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Patentanmeldung Nr. Patent application No. Demande de brevet n°

03076114.2

Der Präsident des Europäischen Patentamts;
Im Auftrag

For the President of the European Patent Office
Le Président de l'Office européen des brevets
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Anmeldung Nr:
Application no.: 03076114.2
Demande no:

Anmelde tag:
Date of filing: 15.04.03
Date de dépôt:

Anmelder/Applicant(s)/Demandeur(s):

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Bezeichnung der Erfindung/Title of the invention/Titre de l'invention:
(Falls die Bezeichnung der Erfindung nicht angegeben ist, siehe Beschreibung.
If no title is shown please refer to the description.
Si aucun titre n'est indiqué se referer à la description.)

Reactor vessel for performing a steam reforming reaction and a process to prepare
synthesis gas

In Anspruch genommene Priorität(en) / Priority(ies) claimed /Priorité(s)
revendiquée(s)

Staat/Tag/Aktenzeichen/State/Date/File no./Pays/Date/Numéro de dépôt:

Internationale Patentklassifikation/International Patent Classification/
Classification internationale des brevets:

B01J/

Am Anmelde tag benannte Vertragstaaten/Contracting states designated at date of
filling/Etats contractants désignées lors du dépôt:

AT BE BG CH CY CZ DE DK EE ES FI FR GB GR HU IE IT LU MC NL
PT RO SE SI SK TR LI

15.04.2003

TS 1327 EPC

Crw

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REACTOR VESSEL FOR PERFORMING A STEAM REFORMING REACTION
AND A PROCESS TO PREPARE SYNTHESIS GAS

Field of the invention

5 The invention is directed to a reactor vessel for performing a steam reforming reaction starting from a natural gas feedstock. The invention is also directed to a process to prepare a mixture comprising carbon monoxide and hydrogen from a carbonaceous feed by performing a partial oxidation reaction and an endothermic steam reforming reaction.

Background of the invention

10 EP-A-168892 describes an endothermic steam reforming reaction, which is carried out in a fixed bed situated in at least one pipe in which a temperature of between 800 and 950 °C is maintained by routing at least part of the hot product gas from a partial oxidation reaction along the pipe(s). According to this publication the combined partial oxidation and endothermic production of synthesis gas result in a better yield of synthesis gas, an increased H₂/CO ratio, a lower usage of oxygen per m³ of synthesis gas product obtained and a lower capital cost of the plant for the production of CO and H₂-containing gas mixtures (as compared to partial oxidation).

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20 25 A reactor and process for performing a steam reforming reaction is described in DE-A-3345088. This publication describes a reactor vessel for performing a steam reforming reaction starting from a natural gas feedstock. The vessel consisted of a tube sheet from which a plurality of tubes filled with a suitable

catalyst extended into the vessel. The required heat of reaction is provided by passing the hot effluent of a partial oxidation reaction of natural gas at the exterior of the reactor tubes in the vessel. Such steam reformer reactors are also referred to as so-called convective steam reformer reactors.

A disadvantage of the known reactor vessel design is that fouling may occur at the exterior surface of the reactor tubes. This fouling will result in a less favourable heat exchange between hot gas and the catalyst bed and in time result in a less efficient operation.

Short run times will result due to frequent shutdowns in order to remove the deposits. Fouling is especially a problem when the hot effluent of a partial oxidation reaction is used. This effluent is especially suited for providing the required heat on the one hand due to its high temperatures. However the soot present in this effluent will cause the above fouling problems and for this reason no commercial applications directed to the combined process of steam reforming and partial oxidation such as described in DE-A-3345088 has been developed according to our knowledge at time of filing.

EP-A-983964 describes a convective steam reforming reactor vessel, wherein the vessel is provided with a plurality of reactor tubes containing a catalyst bed. Around the reactor tubes an annular sleeve is provided to transport a hot effluent of an auto thermal reformer (ATR). By indirect heat exchange between this hot effluent and the reactants passing through the catalyst bed the steam reforming reaction can take place.

WO-A-0137982 discloses a reformer tube of the so-called double-tube configuration of a steam reformer reactor. The double-tube configuration consists of a

reactor tube provided with a catalyst bed in which bed an inner return tube is provided for passage of the reactants being discharged from said catalyst bed. The double-tube configuration is described in more detail in 5 US-A-4690690 according to WO-A-0137982. The inner tube as disclosed in WO-A-0137982 has a non-circular cross-section.

The object of the present invention is to provide a reactor design for a convective steam reformer, which is 10 suitable to make use of a hot gaseous medium, which may cause fouling, as for example the effluent of a partial oxidation, as the heating medium. A further object is to provide a process for the preparation of a mixture of hydrogen and carbon monoxide by a process involving the 15 combination of partial oxidation and steam reforming wherein the problems regarding fouling as described above are minimized.

Summary of the invention

This object is achieved with the following reactor. 20 Reactor vessel for performing a steam reforming reaction comprising:

- a vessel inlet for natural gas and steam,
- a vessel inlet for a hot gaseous medium,
- a vessel outlet for a gaseous product comprising the 25 steam reforming product, and
- a reactor space comprising of a bed of steam reforming catalyst, which reactor space inlet is fluidly connected to the inlet for natural gas and steam and at its outlet end fluidly connected with the outlet for the gaseous product,

30 wherein inside the catalyst bed a passageway is provided fluidly connected to the vessel inlet for the hot gaseous

medium for passage of hot gaseous mixture counter currently to the flow of reactants in the catalyst bed.

The invention is also directed to a process for the preparation of hydrogen and carbon monoxide containing gas from a carbonaceous feedstock by performing the following steps:

(a) partial oxidation of a carbonaceous feedstock thereby obtaining a first gaseous mixture of hydrogen and carbon monoxide and

(b) catalytic steam reforming a carbonaceous feedstock in a Convective Steam Reformer comprising a tubular reactor provided with one or more tubes containing a reforming catalyst, wherein the required heat for the steam reforming reaction is provided by convective heat exchange between the steam reformer reactor tubes and a passageway positioned within and along the axis of the tubular reactor tubes through which passageway the effluent of step (a) flows counter-current to the gasses in the steam reformer tubes.

Step (b) of the above process is preferably performed in the reactor vessel according to the present invention.

Description of the Figures

Figure 1 illustrates a possible embodiment of the reactor according to the invention wherein the catalyst bed is positioned at the shell side of a reactor vessel.

Figure 2 illustrates a steam reformer reactor according to the invention wherein the catalyst bed is contained in one or more reactor tubes. Figures 2a, 2b and 2c illustrate details and preferred embodiments of the reactor of Figure 2.

Figure 3 illustrates a reactor as in Figure 2 wherein two separate gas outlets are provided.

Figure 4 illustrates the combination of the reactor vessel of Figure 3 and a partial oxidation reactor vessel.

5 Figure 5 illustrates an integrated process involving the installation of Figure 4, a Fischer-Tropsch synthesis and some of its downstream unit operations.

Detailed description of the invention

10 Applicants found that by passing the hot gas medium through an elongated passageway gas velocity conditions are achieved wherein soot does not adhere or at least adheres significantly less to the heat exchanging surface of the CSR reactor as compared to the prior art designs. Preferably the elongated passageway is designed such that the gas will flow with a certain velocity wherein the gas has a so-called self cleaning capacity. The gas velocity at design capacity is preferably above 10 m/s and more preferably above 30 m/s. A maximum gas velocity is preferably below 100 m/s and more preferably below 60 m/s. A further advantage is that if any fouling does occur, such deposits could easily be removed in a shut down by state of the art methods for cleaning the interior of conduits. Examples of cleaning methods are pigging and hydrojetting. Further advantages of the above reactor and of its preferred embodiments will be described below.

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Figure 1 illustrates a reactor vessel (1) for performing a steam reforming reaction according to the present invention. Vessel (1) is provided with a vessel inlet (2) for natural gas and steam, a vessel inlet (3) for a hot gaseous medium and a vessel outlet (4) for the gaseous steam reforming product. Between an upper tube sheet (5) and a lower tube sheet (6) a reactor space (7) comprising of a bed (8) of steam reforming catalyst is

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present. Inside the catalyst bed (8) a plurality of passageways (9) are provided which start at tube sheet (5) and extend to a tube sheet (10) positioned below lower tube sheet (6). Through these passageways (9)

5 the hot gaseous medium flows counter currently to the flow of reactants in the catalyst bed (8). The space below this tube sheet (10) defines a space (11) through which the cooled gaseous medium is collected and discharged via vessel outlet (12) from the vessel.

10 Between lower tube sheet (6) and tube sheet (10) a space (13) is provided wherein the natural gas and steam can be distributed and fed to the catalyst bed (8) via openings (15) in tube sheet (6).

15 Preferably the passageways (9) are fixed in the lower tube sheets (6) and are allowed to move relative to tube sheet (5) and (10). This will avoid thermal stress at start-up and shut down operations. The openings may be filled with a flexible seal, for example ceramic fibre, or can be left open as for example illustrated in the 20 aforementioned EP-A-983964. The upper end of the passageways (9) is suitably protected by a heat resistant material, for example a ceramic.

25 Figure 2 illustrates a preferred reactor vessel (20) for performing a steam reforming reaction according to the present invention. The reactor space is defined by one or more parallel positioned reactor tubes (21) filled with a bed (22) of steam reforming catalyst. Through this elongated bed of catalyst one or more passageway(s) (23) for hot gas is provided parallel to the axis of the 30 reactor tube (21). This passageway or conduit (23) allows hot gas medium to exchange its heat by indirect heat exchange through the wall (24) of the conduit (23) with

the counter currently flowing gaseous stream present in the catalyst bed (22).

5 The passageway (23) may have a circular cross-section or a non-circular cross-section such as exemplified in WO-A-0137982. The diameter of a circular passageway (23) is preferably between 10 and 60 mm, more preferably between 20 and 55 mm. The dimensions of a non-circular passageway (23) will be such that the same cross-sectional area results. The inner distance between the 10 outer wall of passageway (23) and the inner wall of reactor tube (21) is preferably between 20 and 150 mm. Preferably this distance is at least six times the size of the catalyst particles which occupy this space.

15 The fact that the catalyst bed (22) is present in an annular space around the passageway (23) is advantageous because less thermal stress will be exercised on the catalyst bed (22) by the tubes (23) at start-up or cool-down situations as would be the case in the reactor of Figure 1.

20 Figure 2 further shows a vessel inlet (26) for natural gas and steam, a vessel inlet for a hot gaseous medium (27) and a vessel outlet (28) for a mixture of the gaseous product comprising the steam reforming product and the hot gaseous medium. Figure 2 also illustrates a 25 tube sheet (29) defining a space (30), which fluidly connects the vessel outlet (28) with the outlet of the passageways (23) extending from said tube sheet (29). Also a tube sheet (31) is present below said tube sheet (29) defining a space (32), which fluidly connects the vessel inlet (26) with the inlet of the reactor tubes (21) filled with catalyst bed (22). The reactor tubes (21) will extend from this tube sheet (31).

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downwards. The passageways (23) will pass this tube sheet up to the upper tube sheet (29) as shown.

In the reactor as shown in part or all of the product gas of the steam reforming reaction is also passed through the elongated passageways. This is advantageous when the hot gaseous mixture is the effluent of the partial oxidation reaction preferably having a steam to carbon ratio of less than 0.5. By combining these streams the steam to carbon ratio of the resulting mixture flowing in the passageways (23) will be higher and thereby minimizing metal dusting corrosion within the passageways (23).

Preferably the lower end of the reactor tube (21) is designed such that immediate and intimate mixing of the steam reformer product which is discharged from the catalyst bed (22) and the hot gaseous medium takes place. Figure 2a shows such a preferred embodiment wherein tube (21) and passageway (23) extend somewhat further than catalyst bed (22). In this extension openings (29) are present in wall (24) such that a substantial amount of the gaseous effluent from the catalyst bed (22) can enter the passageway (23) as is indicated with arrow (29') without being first emitted into the lower end of vessel (20).

Figure 2b shows the upper end of vessel (20) of Figure 2 in more detail. Also shown are the flow directions of the gasses with the arrows. A problem associated with the design shown in Figure 2b is catalyst removal. Between tube sheet (29) and tube sheet (31) a small volume is present for operators to remove catalysts from the catalyst bed (22). The design shown in Figure 2c provides a solution to this problem. Figure 2c shows the same reactor vessel (20) as in Figure 2 except that the

reactor tube (21) extends together with the passageway (23) to the upper tube sheet (29), while the upper level of the catalyst bed (22) extends only to tube sheet (31). In the part of the wall of the reactor tube (21) between tube sheets (31) and (29) an opening (26') is present through which in use gas can enter as shown. In use the reactor tube will be fixed in the upper tube sheet (29) by means of cap (26'') (situation a). By removing the top of vessel (20) the individual reactor tubes/passageways can be removed from the vessel by removing cap (26'') (situation b) and lifting the reactor tube/passageway combination (situation c). Catalyst can then be easily added, removed or refreshed external the reactor vessel (20).

Figure 3 illustrates a special embodiment of the reactor vessel as shown in Figure 2. As in Figure 2 the reactor space is defined by one or more parallel positioned reactor tubes (21) filled with a bed (22) of steam reforming catalyst, comprising a passageway (23) for hot gas. A difference with the reactor of Figure 2 is that a third tube sheet (32) is present at the lower end of the reactor vessel (44) defining a space (33) which fluidly connects the vessel inlet (38) for hot gaseous medium with the inlet of the passageways (23) which penetrate the tube sheet (32) via openings (34) which are preferably larger than the passageway (23) itself. The fact that the passageways are not fixed in the tube sheet (32) is advantageous because it allows the combined reactor tubes (21) and passageways (23) to freely thermally expand in the reactor vessel (44) at start-up and cool down situations. The lower ends of the passageway which extends into the lower space (33) may preferably be made from heat resistant materials like for

example ceramics because of the high temperatures present in said space due to the fact that here the hot gaseous medium enters the reactor via vessel inlet (38).

The outlet opening (35) of the reactor tubes (21) comprising the catalyst bed (22) are positioned just above said tube sheet (32). The steam reforming product being discharged from said opening (35) will enter space (36) defined as the space between tube sheets (40) and (32). This space (36) fluidly connects the vessel outlet (39) for the steam reforming product with the openings (35). The space (36) may suitably be provided with flow directing baffles which will direct the flow of steam reforming product in a zig zag flow through said space thereby optimising the contact of the hot steam reformer product gas and the external surface of the reactor tubes (21) present in said space. In use part of the steam reforming product being discharged from openings (35) will leave the reactor vessel via outlet (39) and part will leave space (36) via openings (34) to space (33) by operating the reactor such that the pressure in space (36) is higher than the pressure in space (33). In use preferably from 0 to 60 wt% and more preferably from 0 to 40 wt% of the steam reformer product, as being discharged from openings (35), may enter space (33) to be mixed with the hot gaseous medium. This is advantageous because the steam to carbon ratio of the gases flowing in the passageways (23) can be increased thereby limiting metal dusting within the passageways (23).

Figure 3 further shows a vessel inlet (43) for natural gas and steam, a vessel inlet for a hot gaseous medium (38), a vessel outlet (39) for steam reforming product and a vessel outlet (42) for the gasses, which

are emitted from the passageways (23). Tube sheets (40) and (41) are present in order to fix the reactor tubes (21) and to define collecting space (45) for the gasses emitted by the passageways (23) and an inlet space (46) fluidly connecting the vessel inlet (43) for steam and natural gas and the reactor tubes (21) comprising the catalyst bed (22).

As in Figure 1 the reactor of Figure 3 yields two product streams. This is advantageous when the hot gaseous medium is the product of a partial oxidation reaction, which has a different hydrogen to carbon monoxide molar ratio than the H₂/CO molar ratio of the steam reformer product. This allows mixing the two streams later on to a specified and desired H₂/CO ratio. It further allows to obtain hydrogen from the stream containing the most hydrogen in a high efficiency. Hydrogen may preferably be separated from such a stream by for example membrane separation followed by a pressure swing absorber step. This is advantageous if the synthesis gas mixture and hydrogen gasses are both required in downstream chemical synthesis processes, for example a Fischer-Tropsch process, wherein synthesis gas is required for the Fischer-Tropsch synthesis reaction and hydrogen is needed for the various hydro-isomerisation/hydrocracking and hydrodewaxing units which convert the FT synthesis product to for example middle distillates and base oils.

It will be evident that the reactor as shown in Figures 1-3 can also be designed up side down and that the variations in tube sheet lay out and corresponding gas inlet and outlet configurations as illustrated are interchangeable between Figures 2 and 3.

On the interior wall of the passageways (23) as illustrated in Figures 1-3 and which are in contact with the effluent of the partial oxidation reaction, carbon may form because part of the carbon monoxide reacts to carbon and carbon dioxide. Also metal dusting corrosion may occur. Furthermore part of the surface may erode resulting eventually in an unacceptable low mechanical integrity of the passageway tubes. These effects are especially significant when the steam to carbon ratio in the hot gas is below 1, more especially below 0.5. Such gas compositions may occur in the above-described embodiments, especially when the partial oxidation of a gaseous hydrocarbon feed is performed in the absence of added steam and/or when the feed to the steam reforming step has a low steam to carbon ratio.

In order further to minimise the above-described coke formation a preferred material for the passageway is suitably used. The passageway (23), suitably in the form of a tube, is preferably made from a metal alloy, wherein the metal alloy comprises from 0 and up to 20 wt% and preferably from 0 up to 7 wt% iron. The alloy preferably also contains between 0 and 5 wt% aluminium, preferably from 0 up to 5 wt% silicon, preferably from 20 up to 50 wt% chromium and preferably at least 35 wt% nickel. Preferably the nickel content balances the total to 100%.

It has been found beneficial to have at least some aluminium and/or silicon in the metal alloy surface when the concentration of steam in the hot gaseous medium within the passageway (23) is lower than 50 vol%, preferably lower than 30 vol% and more preferably lower than 15 vol%. Preferably from 1 up to 5 wt% aluminium and/or from 1 up to 5 wt% silicon is present in said metal alloy under such low steam content conditions. The

resulting aluminium oxide and/or silicon oxide layers will provide an improved protection against coke formation and erosion when the conditions become more reducing at such low steam concentrations. Examples of suitable metals are Inconel 693 containing according to its producer Special Metals Corp (USA), typically comprising 60.5 wt% Ni, 29 wt% Cr and 3.1 wt% Al and the Nicrofer 6025H/6025HT alloys 602/602CA as obtainable from Krupp VDM GmbH (DE).

The invention is also directed to a process for the preparation of hydrogen and carbon monoxide containing gas from a carbonaceous feedstock by performing the following steps:

- (a) partial oxidation of a carbonaceous feedstock thereby obtaining a first gaseous mixture of hydrogen and carbon monoxide and
- (b) catalytic steam reforming a carbonaceous feedstock in a Convective Steam Reformer comprising a tubular reactor provided with one or more tubes containing a reforming catalyst, wherein the required heat for the steam reforming reaction is provided by convective heat exchange between the steam reformer reactor tubes and a passageway positioned within and parallel to the axis of the tubular reactor tube through which passageway the effluent of step (a) flows counter-current to the gasses in the steam reformer tubes.

An advantage of the above process is that due to the fact that the effluent of step (a) is passed through well defined passages through the catalyst bed less fouling on the heat exchanging surfaces of said passageways will occur. A further advantage is that the H_2/CO molar ratio's of the combined synthesis gas products of step (a) and (b) can be from 1.5 up to 3 and even

preferably from 1.9 up to 2.3 making the synthesis gas product suitable for various applications as will be discussed here below.

5 The carbonaceous feedstock in step (a) is preferably a gaseous hydrocarbon, suitably methane, natural gas, associated gas or a mixture of C₁₋₄ hydrocarbons.

10 Examples of gaseous hydrocarbons are natural gas, refinery gas, associated gas or (coal bed) methane and the like. The gaseous hydrocarbons suitably comprises mainly, i.e. more than 90 v/v%, especially more than 94%, C₁₋₄ hydrocarbons, especially comprises at least 60 v/v percent methane, preferably at least 75 percent, more preferably 90 percent. Preferably natural gas or associated gas is used. Preferably any sulphur in the 15 feedstock is removed.

20 Preferably the carbonaceous feed in both steps (a) and (b) is a gaseous feed as described above. In such a preferred embodiment from 10 to 90 wt%, more preferably from 20 to 50 wt%, of the total gaseous feed to steps (a) and (b) is fed to step (b).

25 In step (a) the partial oxidation may be performed according to well known principles as for example described for the Shell Gasification Process in the Oil and Gas Journal, September 6, 1971, pp 85-90. Publications describing examples of partial oxidation processes are EP-A-291111, WO-A-9722547, WO-A-9639354 and WO-A-9603345. In such processes the feed is contacted with an oxygen containing gas under partial oxidation conditions preferably in the absence of a catalyst.

30 The oxygen containing gas may be air (containing about 21 percent of oxygen) and preferably oxygen enriched air, suitably containing up to 100 percent of oxygen, preferably containing at least 60 volume percent

oxygen, more preferably at least 80 volume percent, more preferably at least 98 volume percent of oxygen. Oxygen enriched air may be produced via cryogenic techniques, but is preferably produced by a membrane based process, 5 e.g. the process as described in WO 93/06041.

Contacting the feed with the oxygen containing gas in step (a) is preferably performed in a burner placed in a reactor vessel. To adjust the H₂/CO ratio in the gaseous product obtained in the partial oxidation reaction in 10 step (a), carbon dioxide and/or steam may be introduced into the feed. Preferably up to 15% volume based on the amount of gaseous product, preferably up to 8% volume, more preferably up to 4% volume, of either carbon dioxide or steam is added to the feed. As a suitable steam 15 source, water produced in an optional downstream hydrocarbon synthesis may be used.

The gaseous product of the partial oxidation reaction in step (a) preferably has a temperature of between 1100 and 1500 °C and an H₂/CO molar ratio of from 1.5 up to 20 2.6; preferably from 1.6 up to 2.2.

Step (b) may be performed by well-known steam reforming processes, wherein steam and the gaseous hydrocarbon feed are contacted with a suitable reforming catalyst in a CSR reactor. The catalyst and process 25 conditions as applied in the steam reformer reactor tubes may be those known by the skilled person in the field of steam reforming. Suitable catalysts comprise nickel optionally applied on a carrier, for example alumina. The space velocity of the gaseous feed is preferably from 700 to 1000 litre (S.T.P.)/litre catalyst/hour, wherein. 30 S.T.P. means Standard Temperature of 15 °C and pressure of 1 bar abs. The steam to carbon (as hydrocarbon and CO) molar ratio is preferably from 0 up to 2.5 and more

preferably below 1 and most preferably from 0.5 up to 0.9. If such low steam to carbon ratio's are applied in step (b) the catalyst preferably comprises a Group VIII metal. More preferably the catalyst comprises (a) an 5
oxidic support material and (b) a coating comprising between about 0.1 and about 7.0 wt% of at least one of the metals of the group consisting of Pt, Ni, Pd and Co, preferably platinum; said support material comprising: (i) at least 80 wt% of ZrO_2 which has been calcined at a 10
temperature up to about 670 °C before the application of said coating; (ii) 0.5-10 mol% of at least one oxide selected from the group consisting of Y, La, Al, Ca, Ce and Si, preferably La_2O_3 . Examples of such catalysts are for example the catalyst described in EP-A-695279.

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Preferably the feed also comprises an amount of CO_2 , wherein preferably the CO_2 over carbon (as hydrocarbon and CO) molar ratio is from 0.5 up to 2. The product gas of step (b) preferably has a temperature of from 600 up to 1000 °C and a H_2/CO molar ratio of from 0.5 up to 2.5.

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The temperature of the hydrogen and carbon monoxide containing gas is preferably reduced in step (b) from a temperature of from 1000 up to 1500 °C to a temperature from 300 up to 750 °C. The temperature of the metal wall surfaces of the passageway in step (b) is preferably maintained below 1100 °C.

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The mixture of carbon monoxide and hydrogen (steam reformer product) as obtained in step (b) may be directly combined with the product gas as obtained in step (a). This may be achieved within the reactor as exemplified in Figures 1 and 3. Alternatively the product gas as obtained in steps (a) and (b) may be obtained as separate streams and optionally combined in any desired ratio.

In a preferred embodiment the steam reforming product as obtained in step (b) is fed to step (a). The invention is also directed to encompass below process embodiments wherein a convective steam reactor is used which has 5 separate outlets for the steam reactor product and the cooled effluent of step (a). An advantage of mixing the steam reformer product of step (b) with the feed to step (a) or more preferably directly into the reactor of step (a) is that any methane or higher gaseous 10 hydrocarbon, which may still be present in the steam reformer product, is then further converted to hydrogen and carbon monoxide. This is especially advantageous when the steam reforming step (b) is performed on a feed having a steam to carbon ratio of less than 1, especially 15 between 0.5 and 0.9. Operating the process with a lower steam to carbon ratio in the feed to step (b) is advantageous because the resulting synthesis gas product will then also contain less steam and because smaller 20 reactor equipment may be applied. A disadvantage of operating step (b) at a low steam to carbon ratio is that more unconverted methane will be present in the steam reformer product. By routing the steam reformer product to step (a) this disadvantage is overcome.

25 Optionally the combined streams of step (a) and (b) are subjected to an autothermal reformer step (c) at the elevated temperatures of step (a) in order to achieve convert the gaseous mixture obtained in step (a) to a mixture having a H₂/CO molar ratio closer to the desired thermal equilibrium H₂/CO molar ratio values valid for 30 said operating temperatures. The combined mixture, optionally after performing step (c), is passed via the passageways to provide the required reaction heat for

performing step (b) to yield the synthesis gas end product.

The above embodiment is illustrated in Figure 4. Figure 4 shows the reactor vessel (44) of Figure 3. For clarity reasons no internals of vessel (44) are shown in Figure 4. Also shown is a partial oxidation reactor (51) provided with a burner (52). A carbonaceous feed (50) and an oxygen containing gas (50') is supplied to burner (52). Also shown is that the product gas (55) of step (b) is fed to the upper half of the reactor vessel (51).

Preferably the steam reformer product (55) is fed close, i.e. in the upper half of vessel (51), to the burner (52) in order to benefit the most of the elevated temperatures of from 800 up to 1050 °C present in that region of the vessel (51). Figure 4 also shows a preferred autothermal reformer step (c) when the combined gasses of the partial oxidation reaction and the steam reformer product (55) pass a steam reforming catalyst bed (53) as present in the lower half of reactor vessel (51). The catalyst bed (53) may be any well-known reformer catalyst, for example a Ni-containing catalyst. The effluent of the autothermal reformer step (c) is subsequently fed to inlet (38) of vessel (44), wherein the gasses are cooled in the passageways (23) (not shown in this Figure) and obtained as the final synthesis gas product (63) via outlet (42).

The synthesis gas (63) as obtained by the above process may advantageously be used as feedstock for a Fischer-Tropsch synthesis process, methanol synthesis process, a di-methyl ether synthesis process, an acetic acid synthesis process, ammonia synthesis process or to other processes which use a synthesis gas mixture as feed

such as for example processes involving carbonylation and hydroformylation reactions. To steps (a) and (b) preferably recycle gases are fed. These recycle gasses are obtained in, for example the above exemplified, processes which use the synthesis gas as prepared by the process according to the invention. These recycle gasses may comprise C₁-5 hydrocarbons, preferably C₁-4 hydrocarbons, more preferably C₁-3 hydrocarbons. These hydrocarbons, or mixtures thereof, are gaseous at 5 temperatures of 5-30 °C (1 bar), especially at 20 °C (1 bar). Further, oxygenated compounds, e.g. methanol, 10 dimethylether, acetic acid may be present.

The invention is especially directed to the above process for the preparation of hydrogen and carbon monoxide containing gas (synthesis gas), wherein additional steps (d) (e) and (f) are also performed. In step (d) the synthesis gas is catalytically converted using a Fischer-Tropsch catalyst into a hydrocarbons comprising stream. In step (e) the hydrocarbons comprising stream of step (d) is separated into a hydrocarbon product and a gaseous recycle stream. Suitably the hydrocarbon product are those having 5 or more carbon atoms, preferably having 4 or more carbon atoms and more preferably having 3 or more carbon atoms. The gaseous recycle stream may comprise normally gaseous hydrocarbons produced in the synthesis process, nitrogen, unconverted methane and other feedstock hydrocarbons, unconverted carbon monoxide, carbon dioxide, hydrogen and water.

In step (e) the recycle stream is fed to step (a) and/or (b). Preferably the recycle stream is supplied to the burner of step (a) or directly supplied to the interior of the partial oxidation reactor.

Optionally part or all of the carbon dioxide present in such a recycle stream is separated from said recycle stream before being fed to step (a). Part of the carbon dioxide may suitably be fed to step (a).

5 Step (d) and (e) may be performed by the well known Fischer-Tropsch processes which are for example the Sasol process and the Shell Middle Distillate Process. Examples of suitable catalysts are based on iron and cobalt. Typical reactor configurations include slurry reactors and tubular reactors. These and other processes are for 10 example described in more detail in EP-A-776959, EP-A-668342, US-A-4943672, US-A-5059299, WO-A-9934917 and WO-A-9920720.

15 Figure 5 illustrates the configuration of Figure 4 in combination with a Fischer-Tropsch synthesis process unit (64) and its downstream hydroconversion unit(s) (66). In addition to Figure 4 Figure 5 shows how the synthesis gas (63) is fed to Fischer-Tropsch synthesis process unit (64). In unit (64) a gaseous 20 recycle stream (54) is separated from the hydrocarbon product (65) and recycled to partial oxidation reactor (51). Also shown is how part (60) of the steam reformer product (55) having a relatively high hydrogen over CO molar ratio is fed to a hydrogen recovery unit to 25 obtain hydrogen (62) suitably for use in hydroprocessing unit(s) (66). In said units the hydrocarbon product present in (65), comprising typically a relatively large portion of compounds boiling above 370 °C, is converted by well-known hydrocracking and hydroisomerisation processes to middle distillates. Any remaining residue 30 may be further converted to base oils by catalytic dewaxing processes (not shown), which also require hydrogen. Examples of such downstream hydroprocessing

units are described in for example WO-A-0107538, WO-02070631, WO-02070629 and WO-02070627 and in the references cited in these publications.

5 The following examples will illustrate how the reactor according the invention may be used in a process to make a mixture of carbon monoxide and hydrogen. The values presented are calculated values will come close to the actual values because use has been made of well known thermodynamic relations known to the skilled person in

10 the field of gasification and steam reforming.

Example 1

To a steam reformer reactor according to Figure 2 natural gas and steam are fed in a steam to carbon ratio of 0.75. Also a hot effluent of a partial oxidation reactor is fed via 27 to said reactor. The mass flows, 15 temperatures and resultant product streams are described in Table 1.

Example 2

Example 1 is repeated except that the steam to carbon ratio of the feed to the CSR reactor of Figure 2 was equal to 1.

Example 3

In this example the effluent of the reactor tubes was first fed to the partial oxidation reactor as is illustrated in Figure 4 (line 55). No catalyst bed 53 was present in the reactor 51.

Example 1: Parallel operation - steam to carbon ratio of 0.75

| Stream name | Feed to the partial oxidation reactor | Feed to the reactor of Figure 2 via inlet (26). | Effluent of the reactor entering the reactor of Figure 2 via (27) | Gas leaving the reactor tubes 21 of the reactor of Figure 2 via (26) | Gas leaving the reactor tubes 21 of Figure 26 |
|-----------------|---------------------------------------|---|---|--|---|
| Component | Flow rate kmol/h | Flow rate kmol/h | Flow rate kmol/h | Flow rate kmol/h | Flow rate kmol/h |
| | mol% | mol% | mol% | mol% | mol% |
| | (dry) | (dry) | (dry) | (dry) | (dry) |
| Hexane | 0 | 0.00 | 0 | 0.00 | 0 |
| Pentane | 0 | 0.00 | 0 | 0.00 | 0 |
| Butane | 44 | 0.11 | 13 | 0.12 | 0 |
| Propane | 186 | 0.46 | 56 | 0.52 | 0 |
| Ethane | 1,423 | 3.52 | 427 | 3.94 | 0 |
| Methane | 21,727 | 53.75 | 6,526 | 60.20 | 822 |
| Hydrogen | 613 | 1.52 | 184 | 1.70 | 42,624 |
| Carbon Monoxide | 0 | 0.00 | 0 | 0.00 | 23,604 |
| | | | | | 34.13 |
| | | | | | 7,171 |
| | | | | | 22.39 |

Example 1: Parallel operation - steam to carbon ratio of 0.75

| | | | | | | | | |
|---------------------|--------|--------|--------|--------|--------|--------|--------|--------|
| Carbon | | | | | | | | |
| Dioxide | 267 | 0.66 | | 3,368 | 31.06 | 1,149 | 1.66 | 3,285 |
| Nitrogen | 886 | 2.19 | 266 | 2.45 | 886 | 1.28 | 266 | 10.26 |
| Argon | 76 | 0.19 | 0 | 0.00 | 76 | 0.11 | 0 | 0.83 |
| Oxygen | 15,201 | 37.60 | 0 | 0.00 | 0 | 0.00 | 0 | 0.00 |
| Steam | 0 | 0.00 | 24,543 | 0.00 | 5,034 | 0.00 | 17,538 | 0.00 |
| Total | 40,424 | 100.00 | 35,384 | 100.00 | 74,195 | 100.00 | 49,559 | 100.00 |
| Molecular | | | | | | | | |
| Mass | 22.99 | | 20.32 | | 12.53 | | 14.51 | |
| Temperature (°C) | 353 | | 420 | | 1,273 | | 1,027 | |
| Pressure (bar) | 71 | | 73 | | 70 | | 70 | |

Example 2: Parallel operation - steam to carbon ratio 1.0

| Stream name | Feed to the partial oxidation reactor | Feed to the reactor of Figure 2 via inlet (26) | Effluent of the reactor of Figure 2 via (27) | Gas leaving the reactor tubes 21 of the reactor of Figure 2 via (27) | Gas leaving the reactor tubes 21 of Figure 26 |
|-------------|---------------------------------------|--|--|--|---|
| Component | flow rate | flow rate | flow rate | flow rate | flow rate |
| | kmol/h | kmol% (dry) | kmol/h | kmol% (dry) | kmol/h |
| Hexane | 0 | 0.00 | 0 | 0.00 | 0 |
| Pentane | 0 | 0.00 | 0 | 0.00 | 0 |
| Butane | 45 | 0.11 | 12 | 0.11 | 0 |
| Propane | 191 | 0.46 | 51 | 0.45 | 0 |
| Ethane | 1,456 | 3.52 | 387 | 3.40 | 0 |
| Methane | 22,227 | 53.73 | 5,908 | 51.93 | 822 |
| Hydrogen | 627 | 1.52 | 167 | 1.47 | 43,641 |
| | | | | 61.65 | 19,765 |
| | | | | | 62.74 |

Example 2: Parallel operation - steam to carbon ratio 1.0

| | | | | | | | | |
|------------------|--------|--------|--------|--------|--------|--------|--------|--------|
| Carbon | 0 | 0.00 | 0 | 0.00 | 24,167 | 34.14 | 6,608 | 20.98 |
| Monoxide | 0 | 0.00 | 0 | 0.00 | | | | |
| Carbon Dioxide | 274 | 0.66 | 4,613 | 40.54 | 1,176 | 1.66 | 4,687 | 14.88 |
| Nitrogen | 906 | 2.19 | 241 | 2.12 | 906 | 1.28 | 241 | 0.76 |
| Argon | 78 | 0.19 | 0 | 0.00 | 78 | 0.11 | 0 | 0.00 |
| Oxygen | 15,561 | 37.62 | 0 | 0.00 | 0 | 0.00 | 0 | 0.00 |
| Steam | 0 | 0.00 | 32,573 | 0.00 | 5,151 | 0.00 | 25,816 | 0.00 |
| Total | 41,365 | 100.00 | 43,951 | 100.00 | 75,941 | 100.00 | 57,316 | 100.00 |
| | | | | | | | | |
| Molecular Mass | 22.99 | | 20.62 | | 12.52 | | 15.81 | |
| Temperature (°C) | 353 | | 420 | | 1,273 | | 1,027 | |
| Pressure (barg) | 71 | | 73 | | 70 | | 70 | |

Example 3: Serial operation - steam to carbon ratio 0.75

| Stream name | Component | Feed to the partial oxidation reactor | Feed to the reactor 44 of Figure 4 via inlet 43 | Effluent of the partial oxidation entering the reactor 44 of Figure 4 via inlet 43 | Effluent of the partial oxidation entering the reactor 44 via reactor 44 of Figure 4 via inlet 43 | Gas leaving the reactor tubes of the reactor 44 via reactor 44 via Figure 4 via inlet 43 |
|-----------------|-----------|---------------------------------------|---|--|---|--|
| | | kmol/h | kmol/h | kmol/h (dry) | kmol/h (dry) | kmol/h (dry) |
| | | kmol/h | kmol/h | kmol/h | kmol/h | kmol/h |
| Hexane | 0 | 0.00 | 0 | 0.00 | 0 | 0 |
| Pentane | 0 | 0.00 | 0 | 0.00 | 0 | 0 |
| Butane | 38 | 0.10 | 17 | 0.13 | 0 | 0 |
| Propane | 161 | 0.44 | 73 | 0.55 | 0 | 0 |
| Ethane | 1,232 | 3.34 | 554 | 4.23 | 0 | 0 |
| Methane | 18,815 | 51.06 | 8,453 | 64.55 | 313 | 0.31 |
| Hydrogen | 531 | 1.44 | 239 | 1.82 | 63,382 | 63.39 |
| Carbon Monoxide | 0 | 0.00 | 0 | 0.00 | 30,775 | 30.78 |
| Carbon Dioxide | 232 | 0.63 | 3,415 | 26.08 | 4,323 | 4.32 |

Example 3: Serial operation - steam to carbon ratio 0.75

| | | | | | | | | |
|------------------|--------|--------|--------|--------|---------|--------|--------|--------|
| Nitrogen | 767 | 2.08 | 345 | 2.63 | 1,112 | 1.11 | 345 | 0.89 |
| Argon | 75 | 0.20 | 0 | 0.00 | 75 | 0.08 | 0 | 0.00 |
| Oxygen | 14,998 | 40.70 | 0 | 0.00 | 0 | 0.00 | 0 | 0.00 |
| Steam | 0 | 0.00 | 23,727 | 0.00 | 21,597 | 0.00 | 15,393 | 0.00 |
| Total | 36,851 | 100.00 | 36,822 | 100.00 | 121,577 | 100.00 | 54,287 | 100.00 |
| Molecular Mass | 23.44 | | 20.21 | | 13.23 | | 13.71 | |
| Temperature (°C) | 389 | | 420 | | 1,273 | | 1,027 | |
| Pressure (barg) | 71 | | 73 | | 70 | | 70 | |
| C-atoms | 22,149 | | 13,262 | | 35,411 | | 13,262 | |
| H-atoms | 85,391 | | 85,819 | | 171,210 | | 85,819 | |
| O-atoms | 30,460 | | 30,558 | | 61,018 | | 30,558 | |

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C L A I M S

1. Reactor vessel for performing a steam reforming reaction comprising:
 - a vessel inlet for natural gas and steam,
 - a vessel inlet for a hot gaseous medium,
 - a vessel outlet for a gaseous product comprising the steam reforming product, and
 - a reactor space comprising of a bed of steam reforming catalyst, which reactor space inlet is fluidly connected to the inlet for natural gas and steam and at its outlet end fluidly connected with the outlet for the gaseous product,
wherein inside the catalyst bed a passageway is provided fluidly connected to the vessel inlet for the hot gaseous medium for passage of hot gaseous mixture counter currently to the flow of reactants in the catalyst bed.
2. Reactor according to claim 1, wherein the reactor space is defined by one or more reactor tubes filled with a bed of steam reforming catalyst and wherein said reactor tube comprises one or more passageway(s) running parallel to the axis of said reactor tube.
3. Reactor according to claim 2, wherein the steam reforming product exiting from the reactor tube, the vessel hot gaseous medium from its vessel inlet are both in fluid communication with the inlet of the passageways such that in use a mixture of hot gaseous medium and the steam reforming product passes said passageway.
4. Reactor according to any one of claims 1-3, wherein the passageway is a tube made from a metal alloy, wherein the metal alloy comprises from 0 and up to 7 wt% iron,

between 0 and 5 wt% aluminium, from 0 up to 5 wt% silicon, from 20 up to 50 wt% chromium and at least 35 wt% nickel, wherein the nickel content balances the total to 100%.

5. Process for the preparation of hydrogen and carbon monoxide containing gas from a carbonaceous feedstock by performing the following steps:

10 (a) partial oxidation of a carbonaceous feedstock thereby obtaining a first gaseous mixture of hydrogen and carbon monoxide and

15 (b) catalytic steam reforming a carbonaceous feedstock in a Convective Steam Reformer comprising a tubular reactor provided with one or more tubes containing a reforming catalyst, wherein the required heat for the steam reforming reaction is provided by convective heat exchange between the steam reformer reactor tubes and a passageway positioned within and along the axis of the tubular reactor tubes through which passageway the effluent of step (a) flows counter-current to the gasses

20 in the steam reformer tubes.

6. Process according to claim 5, wherein the gas velocity in the passageway is between 10 and 60 m/s.

7. Process according to any one of claims 5-6, wherein between 0 and 60 wt% of the steam reformer product as obtained in step (b) and the effluent of step (a) flows through the passageway.

25 8. Process according to any one of claims 5-7, wherein the H₂/CO molar ratio of the combined synthesis gas product of step (a) and (b) is between 1.5 and 3, preferably between 1.9 and 2.3.

30 9. Process according to any one of claims 5-8, wherein the steam to carbon molar ratio of the feed to step (b) is between 0.5 and 0.9.

10. Process according to claim 9, wherein the reforming catalyst comprises (a) an oxidic support material and (b) a coating comprising between 0.1 and 7.0 wt% of at least one of the metals of the group consisting of Pt, Ni, Pd and Co, preferably platinum; said support material comprising: (i) at least 80 wt% of ZrO_2 which has been calcined at a temperature up to about 670 °C before the application of said coating; (ii) 0.5-10 mol% of at least one oxide selected from the group consisting of Y, La, Al, Ca, Ce and Si, preferably La_2O_3 .

11. Process according to any one of claims 5-10, wherein the temperature of the metal wall surfaces of the passageways in step (b) is maintained below 1100 °C.

12. Process according to any one of claims 5-11, wherein the steam reforming product of step (b) is fed to step (a).

13. Process according to claim 12, wherein the steam reforming product of step (b) is fed to the upper half of a partial oxidation reactor vessel, said vessel provided with a burner at its upper end, and wherein the temperature in the upper half of the vessel is between 800 to 1050 °C.

14. Process according to any one of claims 12-13, wherein the mixture of the steam reformer product of step (b) and the product of the partial oxidation reaction of step (a) is subjected to an autothermal reformer step (c).

15. Process according to any one of claims 5-14, wherein hydrogen is recovered from the effluent of step (b).

16. Process according to any one of claims 5-15, wherein step (b) is performed in the reactor vessel according to any one of claims 1-4 of the present invention.

17. Process for the preparation of hydrogen and carbon monoxide containing gas from a carbonaceous feedstock by performing the following steps:

(a) partial oxidation of a carbonaceous feedstock

5 thereby obtaining a first gaseous mixture of hydrogen and carbon monoxide and

(b) catalytic steam reforming a carbonaceous feedstock in a Convective Steam Reformer comprising a tubular reactor provided with one or more tubes containing a reforming catalyst, wherein the required heat for the steam reforming reaction is provided by convective heat exchange between the steam reformer reactor tubes and the effluent of step (a) to obtain as separate products a steam reforming product and a cooled effluent of

15 step (a)..

18. Process according to claim 17, wherein the steam to carbon molar ratio of the feed to step (b) is between 0.5 and 0.9.

19. Process according to claim 18, wherein the reforming

20 catalyst comprises (a) an oxidic support material and (b) a coating comprising between 0.1 and 7.0 wt% of at least one of the metals of the group consisting of Pt, Ni, Pd and Co, preferably platinum; said support material comprising: (i) at least 80 wt% of ZrO_2 which has been calcined at a temperature up to about 670 °C before the application of said coating; (ii) 0.5-10 mol% of at least one oxide selected from the group consisting of Y, La, Al, Ca, Ce and Si, preferably La_2O_3 .

25 20. Process according to any one of claims 17-19, wherein the steam reforming product of step (b) is fed to step (a).

30 21. Process according to claim 20, wherein the steam reforming product of step (b) is fed to the upper half of

a partial oxidation reactor vessel, said vessel provided with a burner at its upper end, and wherein the temperature in the upper half of the vessel is between 800 to 1050 °C.

5 22. Process according to any one of claims 20-21, wherein the mixture of the steam reformer product of step (b) and the product of the partial oxidation reaction of step (a) is subjected to an autothermal reformer step (c).

10 23. Process according to any one of claims 17-22, wherein hydrogen is recovered from the effluent of step (b).

15 24. Process according to claim 23, wherein hydrogen is recovered from part of the effluent of step (b), the remaining part of the effluent of step (b) being provided to step (a) or directly combined with the cooled effluent of step (a) and wherein the resultant mixture of carbon monoxide and hydrogen is used in a Fischer-Tropsch synthesis and wherein the hydrogen as prepared is used in one or more hydroconversion process steps in which the Fischer-Tropsch product is further upgraded.

A B S T R A C T

REACTOR VESSEL FOR PERFORMING A STEAM REFORMING REACTION
AND A PROCESS TO PREPARE SYNTHESIS GAS

Reactor vessel for performing a steam reforming reaction comprising:

- a vessel inlet for natural gas and steam,
- a vessel inlet for a hot gaseous medium,
- a vessel outlet for a gaseous product comprising the steam reforming product, and
- a reactor space comprising of a bed of steam reforming catalyst, which reactor space inlet is fluidly connected to the inlet for natural gas and steam and at its outlet end fluidly connected with the outlet for the gaseous product,

wherein inside the catalyst bed a passageway is provided fluidly connected to the vessel inlet for the hot gaseous medium for passage of hot gaseous mixture counter currently to the flow of reactants in the catalyst bed.

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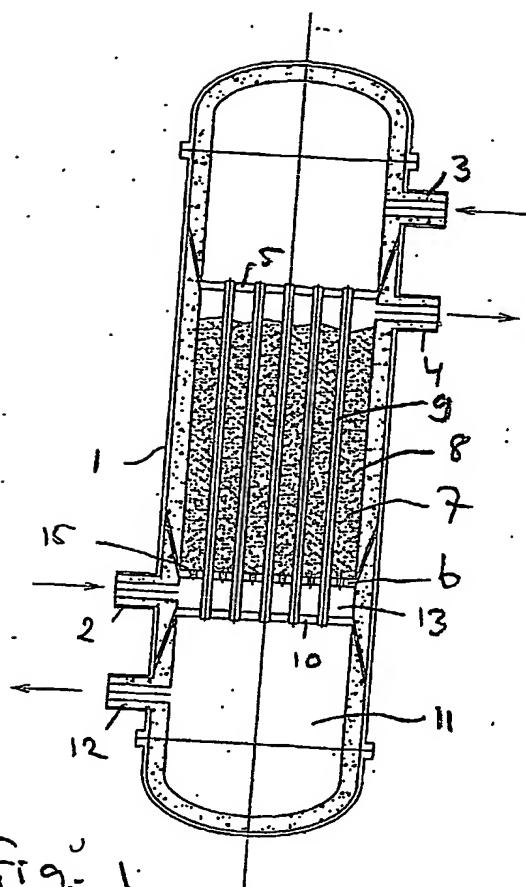


Fig. 1

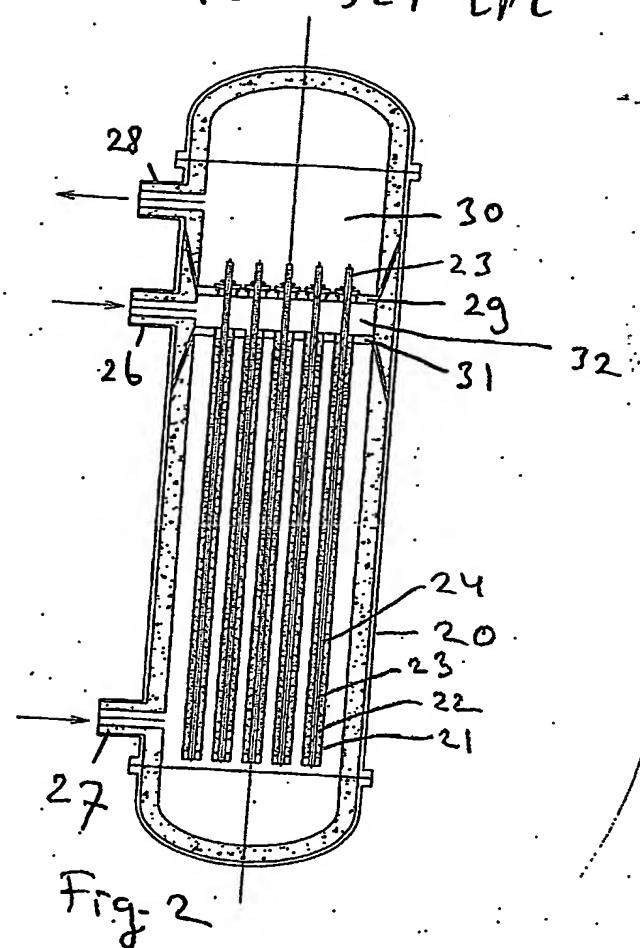


Fig. 2

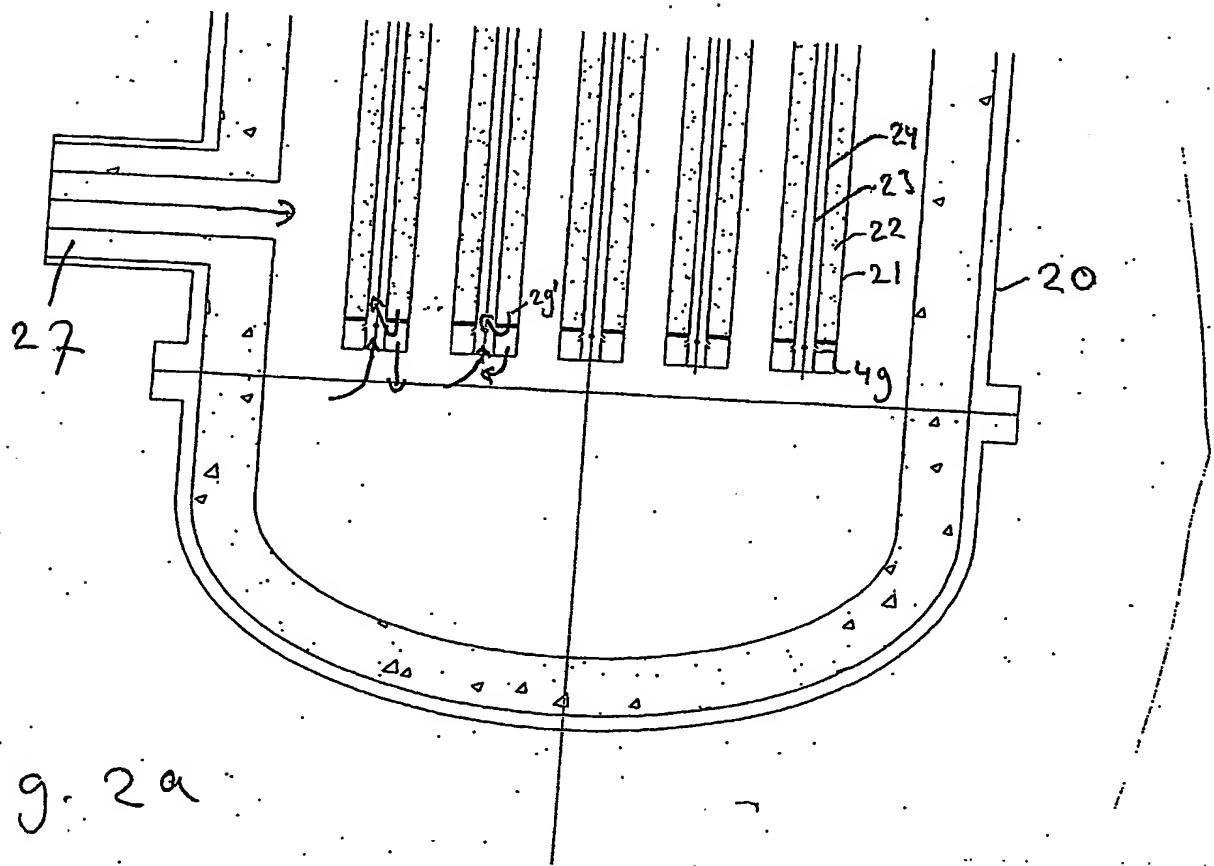


Fig. 2a

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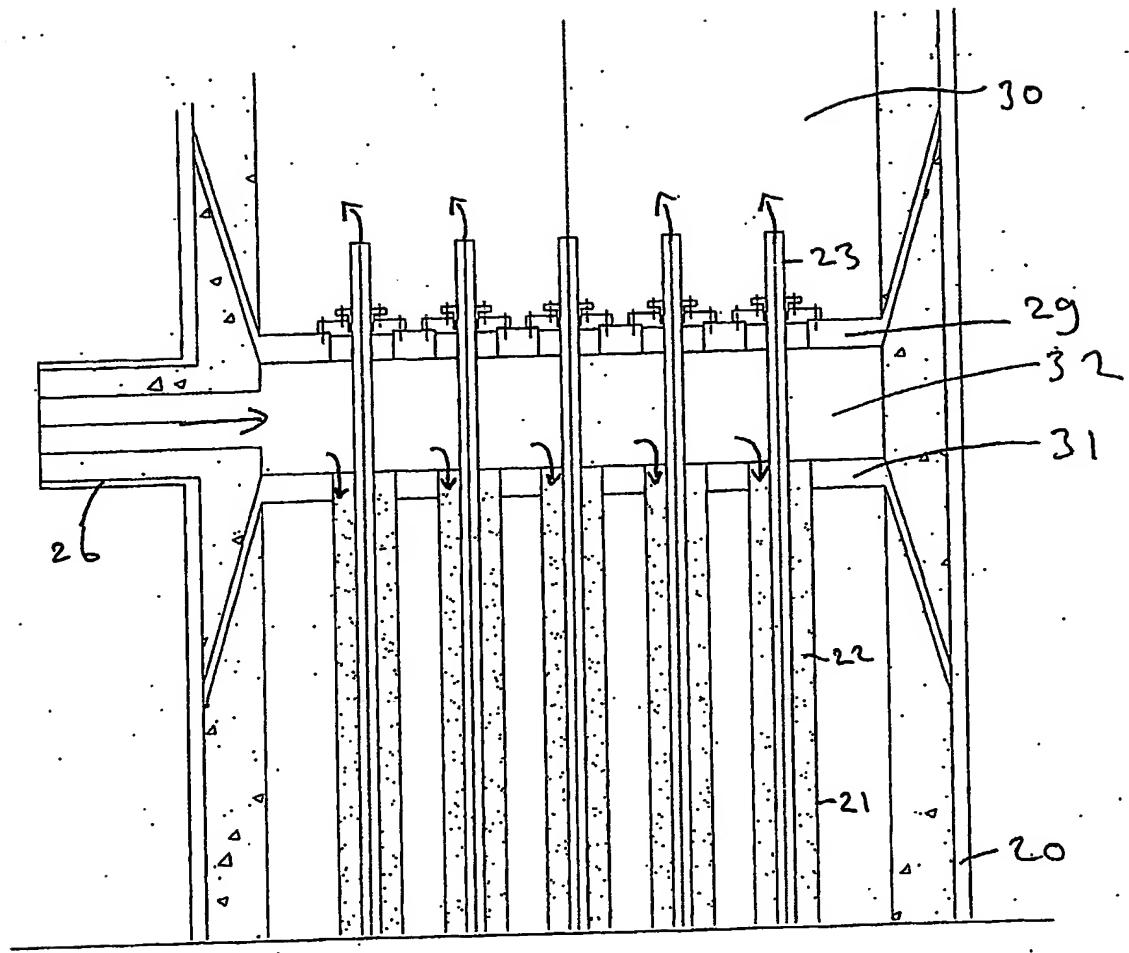


Fig. 2b

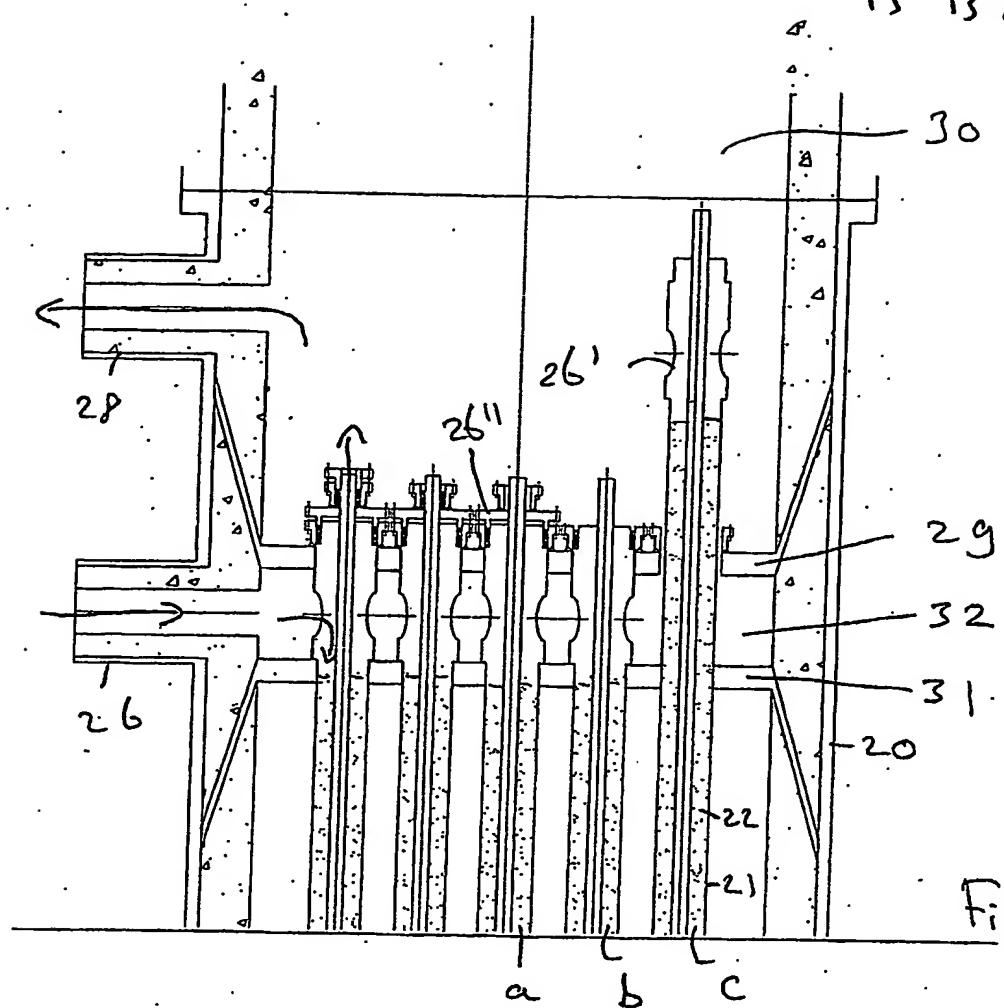


Fig. 2 C

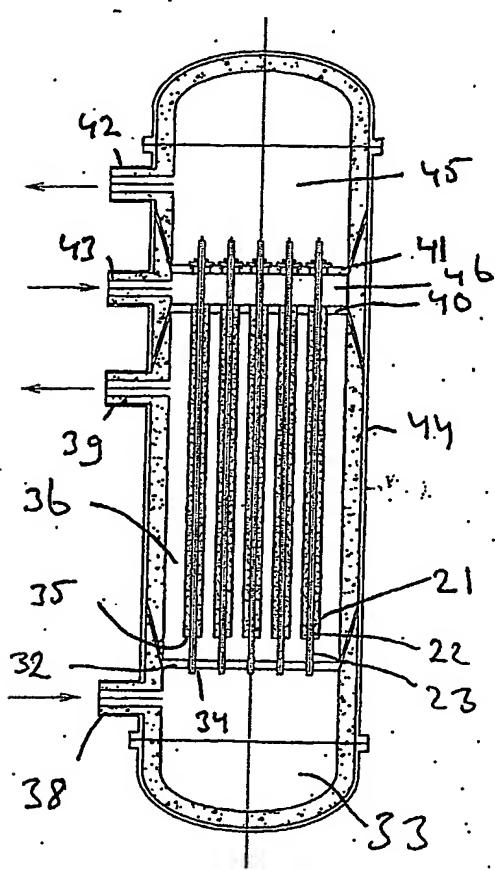


Fig. 3

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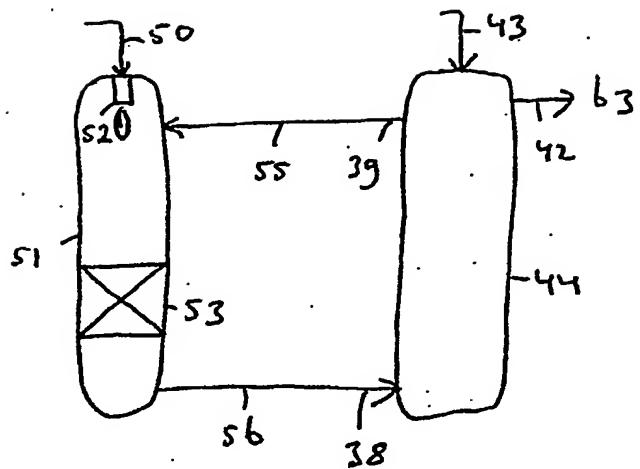


Fig. 4

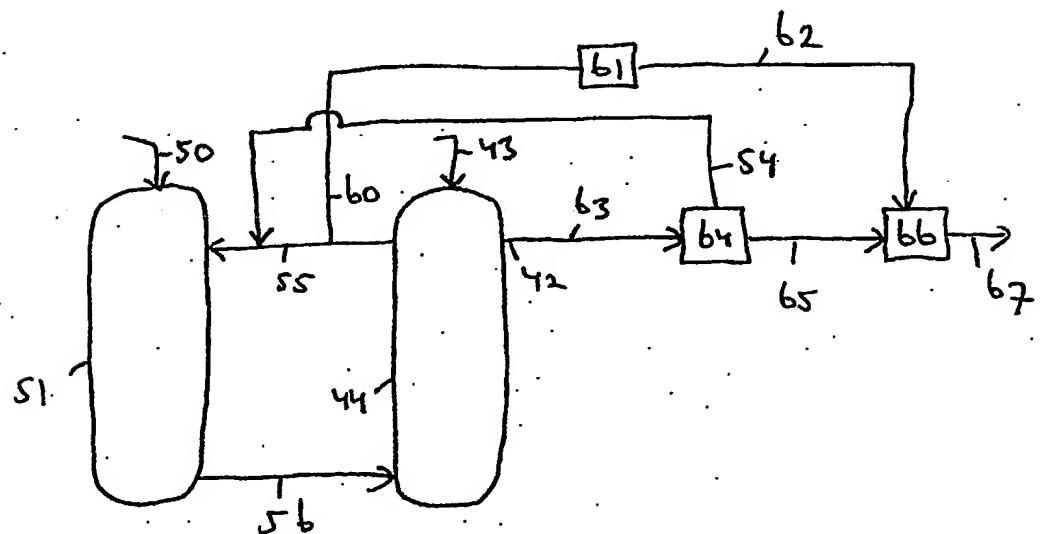


Fig. 5

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